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### (54) Process for catalytic conversion of hydrogen and carbon monoxide

Verfahren zur katalytischen Umsetzung von Wasserstoff und Kohlenmonoxid

Procédé pour la conversion catalytique d'hydrogène et monoxide de carbone

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**EP-A- 0 163 357**                   **EP-A- 0 450 861**

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EP 0 592 176 B1

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**Description**

This invention relates to a hydrocarbon synthesis process comprising reacting a gaseous mixture of hydrogen and carbon monoxide in the presence of a catalyst.

**BACKGROUND OF THE INVENTION**

Slurry reactors are well known for carrying out highly exothermic, three phase, catalytic reactions. Usually called "bubble columns", these reactors have a liquid phase in which solid catalyst particles are dispersed or held in suspension by a gas phase bubbling through the liquid phase. These reactors provide improved heat transfer characteristics for the exothermic reaction, and the bubbling gas provides essentially all of the energy necessary for maintaining the catalyst dispersed in the liquid phase.

These bubble column reactors usually have a multiplicity of vertically arranged tubes suspended within a shell-type housing, the tubes being filled with a heat transfer medium, e.g., steam, which absorbs the heat generated by the exothermic reaction. Such reactors, however, operate at conditions that do not provide high productivity, that is, the rate per unit volume of slurry of CO and H<sub>2</sub> conversion to hydrocarbons, because of their relatively low length to diameter ratios.

Thus, there is a need for a slurry reactor which enhances the reaction rate of the gaseous reactants, efficiently dissipates heat, provides the desirable characteristics of plug flow, and allows for relatively easy replacement of catalyst in the event of declined activity, that is, the rate of conversion of synthesis gas (H<sub>2</sub> + CO) to hydrocarbons is reduced.

EP-A-0450861 discloses a hydrocarbon synthesis process which comprises reacting hydrogen and carbon monoxide at reaction conditions in a slurry bubble column in the presence of catalyst containing cobalt on a titania support or a titania-containing support, and obtaining a relative productivity at least as great as obtained in a plug flow reactor. The reaction conditions of the process may include temperatures of 371.1 to 551.7°C and pressures of 0.5516 to 4.137 MPa.

EP-A-0163357 discloses a reactor comprising a normally vertically extending vessel provided with a plurality of parallel tubes adapted for being filled with catalyst material and substantially vertically extending between an upper tube plate and a lower tube plate, means for circulating a heat transfer medium between the tube plates along the tubes, inlet means above the tubes for distributing liquid in the form of one or more sprays and gas over the plurality of tubes and outlet means below the plurality of tubes for removal of liquid and gas from the vessel, wherein the plurality of tubes are arranged in rows such that the tubes of adjacent rows are staggered, and wherein the upper tube plate is provided with holes, having a lower part in which the

tubes are closely fitted and having a substantially concentric upper part diverging in upward direction, the upper ends of adjacent holes having edges in common in such a manner that the upper end of the upper tube plate is formed by edges arranged in a hexagonal configuration. The reactor is said to be suitable for the synthesis of hydrocarbons from synthesis gas.

The present invention provides a hydrocarbon synthesis process comprising reacting a gaseous mixture of hydrogen and carbon monoxide in the presence of a catalyst within a plurality of vertically-arranged elongated reaction tubes of a reactor of the shell-and-tube type, wherein the said reaction tubes are surrounded by a common heat transfer medium in the shell side of the reactor, characterised in that catalyst is dispersed as a slurry in a liquid in the reaction tubes, and the gaseous mixture is passed upwardly through the slurry in the reaction tubes under conditions of substantial plug flow at a pressure in a range of from 1 to 100 atmosphere (1.014 to 101.38 bar) and a temperature in a range of from 175 to 400°C.

The shell-and-tube type reactor is not unlike a shell-and-tube heat exchanger, the reaction being effected on the tube side and the heat transfer medium being on the shell side.

The energy for maintaining the catalyst dispersed in the reaction tubes may be supplied essentially completely by injection of the gaseous mixture.

Substantially all of the catalyst may be maintained in the reaction tubes.

The catalyst may be a Group VIII metal-containing catalyst, optionally comprising one or more promoters. The catalyst metal (and any promoter(s)) may be supported. The catalyst may be a cobalt-containing catalyst. The cobalt catalyst may be supported.

The process is conducted in apparatus comprising:

- a reactor housing, i.e., the shell;
- more than one, and preferably a multiplicity, of essentially parallel, elongated, vertical reaction zones, i.e., tubes, disposed within the housing and within each zone a slurried catalyst is maintained;
- inlet and outlet means for a heat transfer fluid, liquid or gas or mixture thereof, being in fluid communication and arranged such that the heat transfer fluid is in communication with the external surface of all or substantially all of the reaction zones; that is, the heat transfer fluid is free to flow through the housing between the upper and lower tube sheets and around every reaction tube;
- gas inlet means adapted for receiving feed gas and uniformly dispersing gas bubbles into a liquid, e.g., slurry medium;
- distribution means adapted for reducing bubble size

- or producing adequately small gas bubbles and uniformly distributing the gas bubbles to the reactor tubes;
- slurry liquid inlet means;
  - outlet means for removing the liquid from the reactor housing; alternatively the apparatus can contain liquid product outlet means and separate slurry outlet means; outlet gas means for unconverted feed gas and product gas.

Consequently, as distinguished from the prior art where the process fluid is normally on the shell side of a shell and tube arrangement, the invention described herein maintains the process fluid on the tube side of a shell and tube reactor. Thus, because high length to diameter ratios can now be effected, the advantages for high productivity and high conversion can be obtained while still maintaining the slurry phase beneficial effects of excellent heat transfer and sufficient backmixing in the tubes to enhance selectivity to desired products.

#### BRIEF DESCRIPTION OF THE DRAWINGS

Figure 1 is an elevation of a shell and tube reactor arrangement showing the gas, liquid, catalyst interconnection both above and below the upper and lower tube sheets.

Figure 2 shows a horizontal section A-A through the elevation of Figure 1 showing a possible arrangement of tubes, that is, reaction zones.

Figure 3 is a detail of a possible distribution zone for one reaction tube/zone.

Turning to the drawings for a detailed description of the process and apparatus of this invention, items having the same numbers in different drawings refer to the same items.

Figure 1 is a general layout of the invention where synthesis gas, i.e., hydrogen and carbon monoxide, in a ratio ranging from 1 to 4, preferably, 1.5 to 2.75, more preferably 1.7 to 2.5 is injected into the reactor housing 8 through conduit 12 at superficial gas velocities ranging from about 1 to 20 cm/sec, and through sparger 14 into liquid slurry medium, e.g., hydrocarbon synthesis wax, located in chamber 15. Thus, the gas is injected into an essentially catalyst free zone. The liquid level in the chamber 15 (shown by the wavy line) is maintained by a series of slotted tubes and bubble caps 19 located in distributor tray 18. Note that there is a gas space between distributor tray 18 and the liquid level in chamber 15. This gas space allows essentially even distribution of both gas and liquid into each tube. Without this gas "cushion", a perfectly mixed gas bubble/liquid mixture would be required in chamber 15 to ensure equal distribution to the reactor tubes, thereby requiring significant agitation. This apparatus/design virtually eliminates the need for such energy dissipation in chamber

15.

Above the distributor tray 18 and below tube sheet 22 is a liquid zone allowing interconnection or communication between reaction zones, that may or may not contain catalyst particles. This liquid zone allows uniform mixing of gas and liquid and catalyst and allows even distribution of the liquid, gas and catalyst into the reaction zones. Slurry removal/addition can be made via conduit 32A into the space above distributor tray 18 and below tube sheet 22. The proximity of the tubes 20 to the bubble caps 19 is close enough to allow the process stream (gas, liquid, and catalyst) to move into the tubes but far enough to allow the slurry to move from tube to tube and provide a balance for slurry in each reaction zone. Such a zone will be readily designed by chemical engineers.

There are arranged in the housing 8 a multiplicity of essentially parallel, elongated reaction zones, e.g., tubes 20, which are vertically aligned, that is, aligned along the same longitudinal axis, with the slotted tubes and bubble caps 19. The energy supplied by the injected synthesis gas forces the gas up into the tubes 20 where the gas maintains the catalyst dispersed in the slurry liquid. Reaction takes place wherever there is catalyst and suitable reaction conditions, which include pressures ranging from 1 to 100 atmospheres (1.01 to 101.38 bar), preferably 10 to 50 atmospheres (10.14 to 50.69 bar), more preferably 15 to 40 atmospheres (15.21 to 40.55 bar) and temperatures ranging from 175°C to 400°C, preferably from 180°C to 280°C, more preferably from 180°C to 240°C.

Liquid recycle, for mixing purposes or for supplying additional energy for maintaining the catalyst dispersed is not essential, but may be supplied if desirable through line 16.

The slurry phase liquids in which the catalyst is dispersed are those that are liquid at reaction conditions, generally inert, and a good solvent for synthesis gas. Typically, the slurry is the product of the reaction and contains C<sub>5</sub>+ hydrocarbons, usually C<sub>5</sub>-C<sub>50</sub> hydrocarbons. Preferably, however, the slurry liquid comprises primarily high boiling paraffins with small amounts of primary and secondary alcohols, acids, esters, or mixtures thereof. Sulfur, nitrogen, phosphorus, arsenic, or antimony heteroatoms are to be avoided since these tend to poison the hydrocarbon synthesis catalyst. Examples of specific slurry liquids are dodecane, tetradecane, hexadecane, octadecane, tetracosane. Preferred slurry materials are Fischer-Tropsch waxes and C<sub>16</sub>-C<sub>18</sub> hydrocarbons.

The concentration of solids, including catalyst, in the slurry phase is usually 10-50% by weight, preferably 30-40 wt% solids.

The hydrocarbon synthesis reaction is highly exothermic and the heat of reaction is removed by a heat transfer material 25 on the shell side of the shell and tube arrangement. The common heat transfer material can be any material having a high heat capacity,

whether or not undergoing a phase change. Preferably, the heat transfer fluid is boiling water, and shell side cooling water enters the housing 8 through inlet 24 and exits either as boiling water or as steam through outlet 26. The heat transfer fluid is free to contact the external surfaces of all or substantially all of the reaction zones.

The upper portions of the tubes 20 are secured in an upper tube sheet 28, the length of the tubes or reaction zones being sufficient to provide an L/D ratio of at least about 10, preferably at least about 20, and most preferably at least about 40.

There is a liquid space above the upper tube sheet and from which catalyst slurry addition or removal may be made through line 32. This liquid space allows interconnection or communication between the upper ends of reaction zones for gas, liquid, and catalyst. Above the liquid space is another tube sheet 30 holding filter cartridges 31 which may contain sintered metal mesh, woven metal fibers, glass fibers, cloth, fibrous carbon that can remove the catalyst particles while allowing passage of the liquid. The filter cartridges are each vertically aligned with each reaction zone and prevent catalyst particles from reaching the upper portion of the housing 8. Above the filter cartridge tube sheet is a gas-liquid disengagement zone 36 topped by a foamy interface 38. Liquid product from the hydrocarbon synthesis may be removed via line 40, or alternately via line 33 at or above the filter cartridge tube sheet. A demistor 41 finally separates gas from liquid droplets and residue gases are withdrawn via line 42. Thus, the liquid space above the upper tube ends and below the filter tube sheet allows fluid communication between the upper tube ends and the alternate slurry addition/removal conduit 32 as well as the space above the filter tube sheet. The space above the filters and filter tube sheet allows fluid communication of the gas outlet means, the liquid outlet means and the space below the filter tube sheet, thereby further allowing fluid communication to the upper ends of the reaction tubes.

Figure 2 is a horizontal section through about the center of the housing and shows the reactions zones 20 arranged in the housing 8 and surrounded by a common heat transfer medium 25.

Figure 3 is a proposed detail of the distribution zone for a reaction zone/tube. By virtue of energy supplied by the synthesis gas feed, sparged, bubbly liquid in area 9 enters the slotted bubble cap tube, where the tube is 19A and the bubble cap is 19B. The area above and outside the slotted tube 19A entry is a vapor space 11. The bubble cap/tube 19 is aligned vertically with the tube/reaction zone 20. The bubble cap 19B is operatively connected to the tube 19A which is about 1/2-3 inches (1.27 to 7.62 cm) diameter, the bubble cap is sized to give a pressure drop and injection velocity sufficient to decrease the size of the bubbles and suspend catalyst particles 21 in that reaction zone. In particular, the bubble cap is 1/2 to 4-1/2 inches (1.27 to 11.43 cm) in diameter and the injection velocity at this point is 20

to 100 ft/sec (6.096 to 30.48 m/s).

Thus, not only is bubble size decreased vis-a-vis bubbles coming out of the sparger but an agitated flow is also created. The flow is sufficient to lift and disperse catalyst particles 21, that are in the liquid zone above the distributor 18 and the lower tube sheet 22, into the reaction zone/tube 20. The arrangement and sizing of the bubble caps allow for relatively small, substantially uniform bubbles such that gas/liquid mass transfer rates are enhanced, thereby enhancing catalyst life and CO conversion. The gas to liquid mass transfer rates are at least about equal to the conversion rate of synthesis gas with minimal loss in reaction driving force due to mass transfer resistances. Minimizing such resistances reduces total pressure requirements and achieves high productivity. Lower pressure also reduces equipment and operation expenses.

From the arrangement shown in the drawings, the slurred catalyst is preferably maintained within the reaction tubes although some catalyst may optionally be present in the liquid zones or plenum above and below the upper and lower tube sheets, leading to the interconnection (or communication between) of the reaction tubes for gas, liquid, and slurry.

During operation each reaction zone/tube operates essentially independently of the other reaction zones/tubes even though they are interconnected for all three phases. The rates of gas, liquid, and solid flows through each tube are metered by having a liquid space below the lower tube sheet and above the distributor tray allowing communication between tubes.

Catalyst dispersion in the reaction zones is maintained essentially by energy supplied by the injected and rising synthesis gas. Other means are known for providing energy catalyst dispersion, e.g., slurry recycle, liquid recycle, but it is preferred to provide the energy substantially from injected synthesis gas, i.e., at least 50% thereof, preferably at least 90%, more preferably at least 95% thereof.

The catalyst employed in the hydrocarbon synthesis process is any catalyst known to be active in Fischer-Tropsch synthesis. For example, Group VIII metals, whether supported or unsupported, are known Fischer-Tropsch catalysts. Of these, iron, cobalt and ruthenium are preferred, particularly iron and cobalt, most particularly cobalt.

A preferred catalyst is supported on an inorganic refractory oxide selected from Groups III, IV, V, VI, and VIII of the Periodic chart of the elements. Preferred supports include silica, alumina, silica-alumina, the Group IVB oxides, most preferably Group IVB oxides, specifically titania (primarily in the rutile form), and generally supports having a surface area of less than about 100 m<sup>2</sup>/gm, preferably less than about 70 m<sup>2</sup>/gm.

The catalytic metal is present in catalytically active amounts, usually from 1-50 wt%, preferably 2-40 wt%, more preferably from 2-25 wt%. Promoters may be added to the catalyst and are well known in the Fischer-

Tropsch catalyst art. Promoters can include ruthenium (when it is not the primary catalytic metal), rhenium, hafnium, cerium, and zirconium, and are usually present in amounts less than the primary catalytic metal (except for ruthenium which may be present in co-equal amounts), but the promoter:metal ratio should be at least about 1:10. Preferred promoters are rhenium and hafnium.

Catalyst preparation can be any of a variety of methods including impregnation, spraying and particularly the method described in U.S. Patent US-A-4,621,072 incorporated herein by reference.

Catalyst particle size is not critical and particle sizes may range from that which is reasonably filterable to that which can be reasonably dispersed in a slurry phase. Particle sizes of 1-200 microns (1 - 200  $\mu\text{m}$ ), preferably 20-150 microns (20-150  $\mu\text{m}$ ) meet these requirements.

#### Example

A bubble column reactor consisting of a 6 inch (15.24 cm), Schedule 80 stainless steel pipe about 50 feet (15.24 m) in length was mounted inside a 12 inch (30.48 cm) pipe that served as the cooling jacket. The internal diameter of the reactor was 5.76 inches (14.63 cm). Feed gas to the reactor was preheated and injected through a single, 1/2 inch (1.27 cm) nozzle at the bottom of a conical inlet section. The height of the gas-liquid-solid slurry varied from about 12 feet (3.658 m) to about 32 feet (9.754 m). An axial thermocouple rake\* provided a definition of the reactor temperature profile. A gas jet, installed just above the normal liquid level of the slurry, controlled foam formation and allowed exit gas velocities up to 7.9 cm/sec to be used.

The cobalt containing catalyst was prepared by impregnation of titania extrudates, which were then crushed and screened to provide the desired particle size of about 30 microns (30  $\mu\text{m}$ ) mean diameter. The catalyst was activated in the reactor, before a slurry wax liquid was introduced, by passing hydrogen upflow through the fluidized solids.

Starting at a low feed ratio of 1.7 H<sub>2</sub>/CO and a gas superficial velocity of 3.5 cm/s (about 2200 GHSV), the unit lined out quickly, achieving a steady CO conversion level in excess of 50% (then at a 2800 GHSV) within four hours. After a slight drop in conversion to about 45% in the first 14 hours, the feed ratio was increased to 2.1 (at constant feedrate) to prevent any H<sub>2</sub> depletion-related deactivation that might have been occurring. The unit responded with a dramatic increase in conversion, up to about 58%, quickly lining out at about 54%.

Following a unit upset (overhead carryover induced by increasing feedrates) and a brief shutdown, the unit was restarted, this time at a higher gas feedrate (5 cm/sec), a higher steam jacket pressure (140 psig, 10.67 bar.abs.), and a higher reactor pressure (285 psig 20.67 bar.abs.) at the reactor gas outlet). Conversion

promptly reached 58%, then continued to increase to 65%+. The unit maintained extremely high conversion levels, in the range of 60% to 70%. Over the range of space \* a "rake" is a line of thermocouples extending the length of the reactor. velocities tested, from 2800 to 3600 ( $\text{hr}^{-1}$ ), CO productivities ranged from 450 to 750 vols. CO/vol. catalyst/hour for this very active HCS catalyst. After nearly seven days on synthesis, there was no significant activity loss and methane selectivities were low (2.7% to 3.2%). Filtration of the hydrocarbon synthesis liquid (wax) was easy at flux rates in the range of 0.13 to 0.2 gpm/ft<sup>2</sup> (5.297 to 8.149  $\text{l}/\text{min}/\text{m}^2$ ). After 120+ hours on the first post-blowback cycle, the 0.175 ft<sup>2</sup> (0.01626 m<sup>2</sup>), 18 micron (18  $\mu\text{m}$ ), Regimesh® filter showed no indication of needing a second backflush.

#### Claims

1. A hydrocarbon synthesis process comprising reacting a gaseous mixture of hydrogen and carbon monoxide in the presence of a catalyst within a plurality of vertically-arranged elongated reaction tubes of a reactor of the shell-and-tube type, wherein said reaction tubes are surrounded by a common heat transfer medium in the shell side of the reactor, characterised in that catalyst is dispersed as a slurry in a liquid in the reaction tubes, and the gaseous mixture is passed upwardly through the slurry in the reaction tubes under conditions of substantial plug flow at a pressure in a range of from 1 to 100 atmosphere (1.014 to 101.38 bar) and a temperature in a range of from 175 to 400°C.
2. The process of claim 1 wherein the energy for maintaining the catalyst dispersed in the reaction tubes is supplied essentially completely by injection of the gaseous mixture.
3. The process of claim 1 or claim 2 wherein substantially all of the catalyst is maintained in the reaction tubes.
4. The process of any one of claims 1 to 3 wherein the catalyst is a Group VIII metal containing catalyst, optionally comprising one or more promoters.
5. The process of claim 4 wherein the catalyst metal (and any promoter(s)) is supported.
6. The process of claim 4 or claim 5 wherein the catalyst is a cobalt-containing catalyst.
7. The process of any one of claims 4 to 6 wherein the cobalt catalyst is supported.

### Patentansprüche

1. Kohlenwasserstoffsyntheseverfahren, bei dem eine gasförmige Mischung aus Wasserstoff und Kohlenmonoxid in Gegenwart eines Katalysators innerhalb einer Vielzahl von senkrecht angeordneten länglichen Reaktionsrohren eines Reaktors vom Rohrbündel-Typ umgesetzt wird, wobei die Reaktionsrohre von einem gemeinsamen Wärmeübertragungsmedium in der Schalenseite des Reaktors umgeben sind, dadurch gekennzeichnet, daß Katalysator als Aufschämmung in einer Flüssigkeit in den Reaktionsrohren dispergiert ist und die gasförmige Mischung durch die Aufschämmung in den Reaktionsrohren unter Bedingungen von im wesentlichen Propfenfluß bei einem Druck im Bereich von 1 bis 100 Atmosphären (1,014 bis 101,38 bar) und einer Temperatur im Bereich von 175 bis 400 °C aufwärts geführt wird. 5
  2. Verfahren nach Anspruch 1, bei dem die Energie, um den Katalysator in den Reaktionsrohren dispergiert zu halten, im wesentlichen vollständig durch die Einspritzung der gasförmigen Mischung geliefert wird. 10
  3. Verfahren nach Anspruch 1 oder Anspruch 2, bei dem im wesentlichen der ganze Katalysator in den Reaktionsrohren gehalten wird. 15
  4. Verfahren nach einem der Ansprüche 1 bis 3, bei dem der Katalysator ein Katalysator ist, der Metall der Gruppe VIII enthält und gegebenenfalls einen oder mehrere Beschleuniger (Promoter) enthält. 20
  5. Verfahren nach Anspruch 4, bei dem das Katalysatormetall (und jeglicher (jegliche) Beschleuniger)) getragen ist. 25
  6. Verfahren nach Anspruch 4 oder Anspruch 5, bei dem der Katalysator ein Kobalt enthaltender Katalysator ist. 30
  7. Verfahren nach einem der Ansprüche 4 bis 6, bei dem der Kobaltkatalysator getragen ist. 35
- 40
- 45

### Revendications

1. Procédé de synthèse d'hydrocarbures dans lequel on fait réagir un mélange gazeux d'hydrogène et de monoxyde de carbone en présence d'un catalyseur dans une série de tubes réactionnels allongés, agencés verticalement, d'un réacteur du type à coque-tubes, dans lequel lesdits tubes réactionnels sont entourés d'un agent de transfert thermique commun sur le côté coque du réacteur, caractérisé en ce que le catalyseur est dispersé sous la forme d'une suspension dans un liquide dans les tubes 50
2. Procédé selon la revendication 1, dans lequel l'énergie permettant de maintenir le catalyseur dispersé dans les tubes réactionnels est fournie de manière sensiblement intégrale par injection du mélange gazeux. 55
3. Procédé selon la revendication 1 ou 2, dans lequel sensiblement la totalité du catalyseur est maintenue dans les tubes réactionnels.
4. Procédé selon l'une quelconque des revendications 1 à 3, dans lequel le catalyseur est un catalyseur contenant un métal du groupe VIII et comprenant en option un ou plusieurs promoteurs.
5. Procédé selon la revendication 4, dans lequel le métal catalytique (et le ou les promoteurs) est supporté.
6. Procédé selon la revendication 4 ou 5, dans lequel le catalyseur est un catalyseur contenant du cobalt.
7. Procédé selon l'une quelconque des revendications 4 à 6, dans lequel le catalyseur de cobalt est supporté.

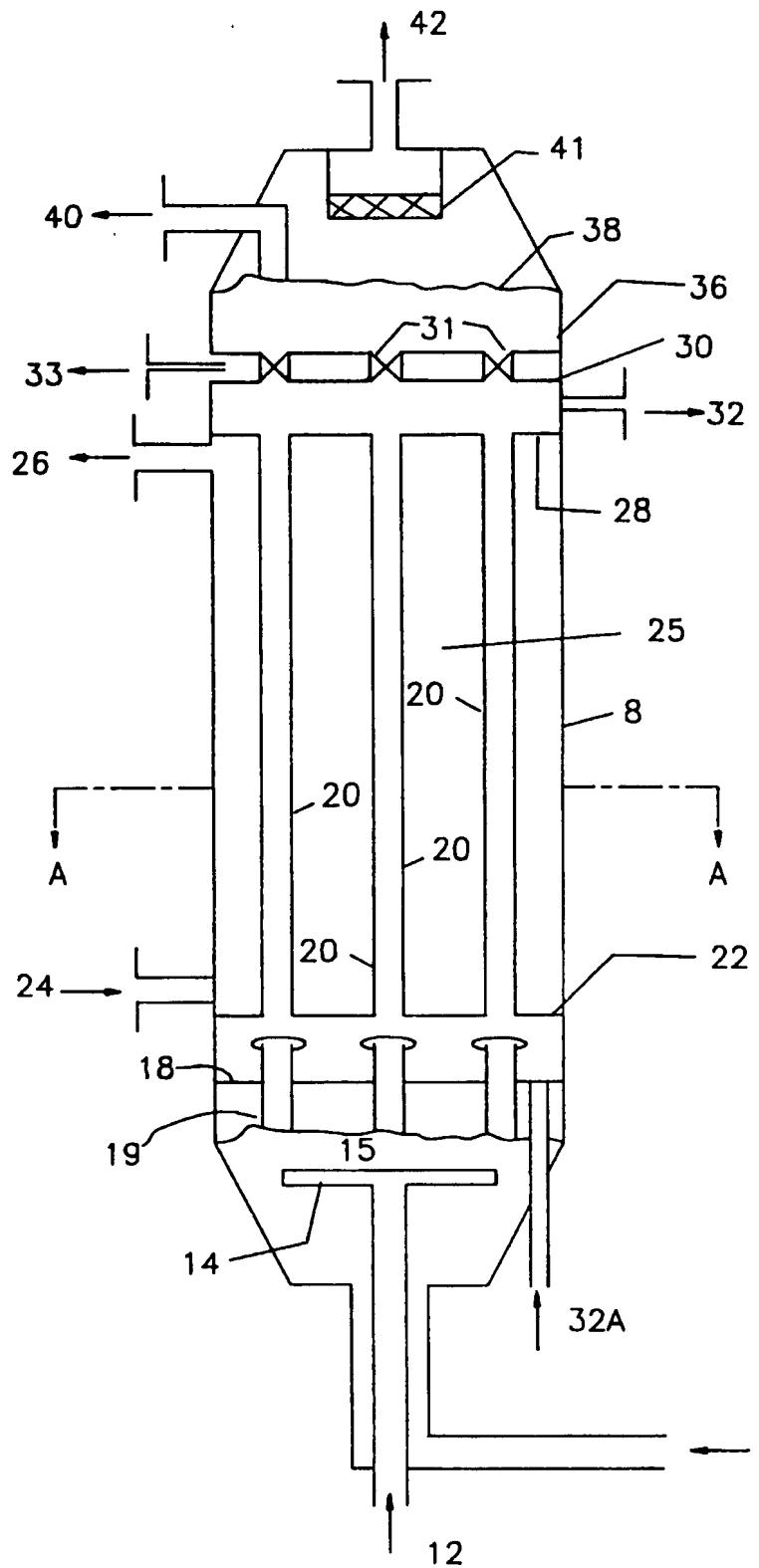


FIG. 1

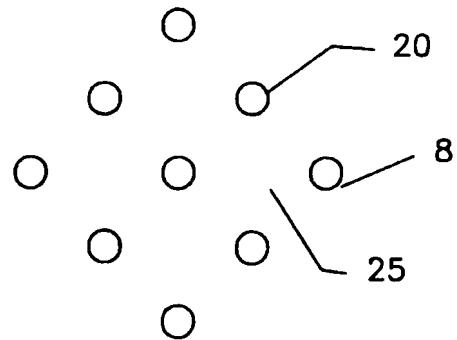


FIG. 2

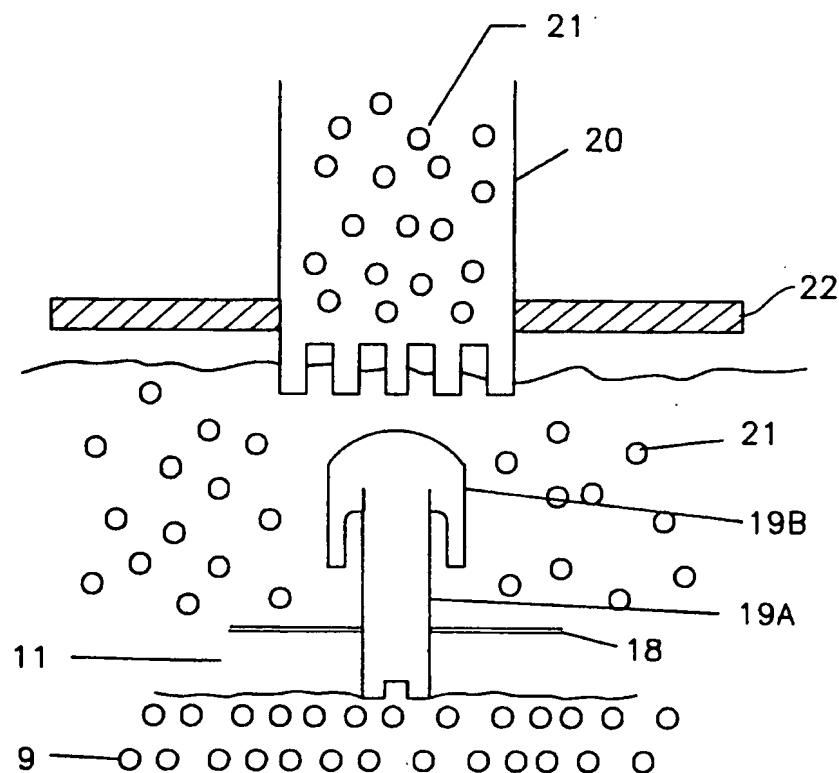


FIG. 3